Value Creation of Liquefied Petroleum Gas (LPG)

by

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Dissertation submitted in partial fulfillment of

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(Chemical Engineering)

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CERTIFICATION OF APPROVAL

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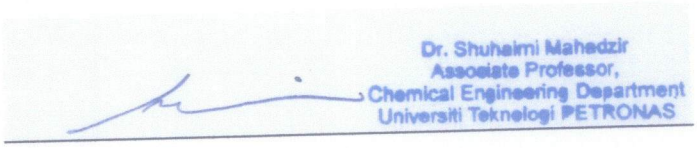
Approved by,



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UNIVERSITI TEKNOLOGI PETRONAS

TRONOH, PERAK

September 2013

CERTIFICATION OF ORIGINALITY

This is to certify I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

JEREMY STA MARIA

ABSTRACT

LPG constitutes a mixture of propane, normal butane and isobutane that are very valuable not just as a source of fuel but also as petrochemical feedstock to produce numerous products such as plastics and synthetic rubbers. Thus, if they are purified or separated, the market value of the constituents is higher as compared to their mixture in LPG. Isobutane especially, has the highest market price among the 3 components in LPG. In spite of the potential value mentioned, to get a high purity of the individual components, separation process will be required. In this case, separation of multicomponent will require at least 2 distillation columns. Low relative volatility, in the case for the separation of isobutane and normal butane, will require large number of stages for their separation. As the number of trays for the column gets bigger, the capital expenditure for the column will increase. Therefore, it is important that the process of purifying is optimally traded off between the operating costs and capital expenditure. Using Aspen HYSYS simulation software, the base case simulation of the plant was simulated. After identifying potential aspect that can be manipulated in the processes for optimization, they will be simulated. The base case results were used as the benchmark for evaluation against the alternative separation process simulations that have been successfully simulated namely, dividing wall column (DWC) and indirect sequencing. This study has shown that the direct sequence (base case), indirect sequence and DWC produces product value of RM 41 000 to RM 42 000 per hour. Results from this simulation have indicated that, the direct sequence is preferred as its throughput is the highest with 1.267 kg of products purified per kW of energy used as compared to the indirect and DWC case with throughputs of 1.085 and 0.316 of purified constituents per kW of energy consumed. The outcome of this study has been achieved while providing an insight on the preliminary analysis of feasibility of this study of creating added value from LPG.

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CHAPTER 1

INTRODUCTION

1.1 Background of Study

1.1.1 Liquefied Petroleum Gas (LPG)

Liquefied petroleum gas (LPG), a product refined from natural gas or crude oil, is primarily used as domestic fuel for cooking, heating and automobile fuel. It comprises a mixture of butane to propane at ratio, typically at 7030 by mole fraction, subject to country specifications, weather or manufacturer. In Malaysia, the LPG composition that is bottled contains mainly of propane and butane at ratio of 6040 by mole fraction (refer to Table 2) (Suruhanjaya Tenaga, 2008). In spite of that, the presence of trace amount of compounds such as ethane and pentane is common.

Petroleum refining and natural gas production yield LPG as their by-product. In petroleum refining, LPG is produced in various stages such as atmospheric distillation, reforming and cracking processes. According to World LP Gas Association, production of LPG from natural gas depends on the "wetness" of the natural gas produced which account about 1 to 10% of the unprocessed gas stream, much more than crude oil refining which contributes between 1 to 4% only (World LP Gas Association, 2006). In total, the amount of LPG production, worldwide, that originates from natural gas is about 60% and remainder 40% is from crude oil refining (Sibbald, 2010).

For industrial application, LPG is used as feedstock to create olefins such as ethylene and propylene. Olefins can further be polymerized to produce polyethylene, polypropylene and other higher value products (Richardson, 2010). Economic activity and the rising need for plastics will influence the supply and demand of LPG consequently; it will affect the price of LPG globally (PennWell, 2006). In spite of that, in Malaysia, the price is regulated by the government therefore; the price of LPG remains fixed under the government's price control mechanism. In Malaysia, the price of bottled LPG is RM 1.85 per kg (MIDA, 2012).

The demand for LPG in Asia is still far behind as compared to Europe. Quoting Paul Hodges, Chairman of United Kingdom-based chemical consultancy, European countries have utilized LPG for more applications as compared to Asian countries. Asians are still discovering and currently, investing to increase the application of LPG. In 2010, the demand for LPG in Europe was about 39.2 million tonnes and the Southeast Asia, combined with Pacific (including Australia) was just about 16.1 million tonnes (Hart, Gist, & Otto, 2011). However, quoting ICIS News, it is forecasted that the demand for isobutane in China is expected to surge for numerous deep processing units (Lv, 2013). According to ICIS news agency, the apparent increase in terms of demand of LPG in China is expected to increase 4.81% yearly for numerous deep-processing units that will come into operation next year. In the Southeast Asian region, the growth in demand was mainly due to Malaysia and Thailand which represent 60% of the total demand in the region (PennWell, 2006).

The potential in creating additional value for LPG by extracting the different compounds of C_4 hydrocarbon mixtures, namely, isobutane and n-butane that is in the LPG has to be evaluated. In order to extract them, the process of distillation would be the subject that will be studied in order to optimize potential processes for the separation of isobutane and n-butane in order to get the most value out of the products separated instead of burning them directly. This also has to be weighed against the sales margin of the products. This will further be discussed in the later chapters.

LPG is claimed to be green which means LPG is more environmental friendly compared to other fuels as it emits 50% less CO_2 emissions than coal and 20% less than heating oil (World LP Gas Association, 2006). Aside from that, World LP Gas Association (2006) stated that LPG produces "50% less carbon monoxide, 40% less hydrocarbons, 35% less nitrogen oxides (NO_x) and 50% less ozone forming potential as compared to gasoline". The properties of LPG which requires an LPG to air ratio of 2 to 9% for ignition should also be considered as another advantage in terms of safety as the concentration for ignition is difficult to attain (Hindustan Petroleum Corporation Limited, 2008). When compared to other fuels, LPG has a high calorific value of 11Kcal/kg which will provide a high content of heat available when LPG is completely burned or it can also be said to be more efficient fuel than those of lower calorific value. Refer to Table 1 for the caloric value of LPG as compared to other fuels.

Table 1: Calorific value comparison with other fuels
(Source Hindustan Petroleum web site)

Fuel	KCal / Kg(At room temperature)	Heat transfer efficiency
LPG	11900	85%
SKO	11100	50%
Light Diesel Oil (LDO)	10700	60%
Furnace Oil (FO)	10280	55%
Firewood	4400	15% to 20%
Electricity	860/Kw	65%

The main factor which affects the percentage of propane and butane in LPG is the weather of the country. During the cold season, LPG is often filled with more propane than butane and vice versa during the hot season. The reason propane is filled more during lower temperature is because of its lower boiling point which is -42°C (-44°F) compared to butane which is approximately -0.6°C (31°F) (Ophardt,

2003). At very cold temperature, the boiling point of butane restricts it from vaporizing, thus, limiting its usage. That is why propane is mixed with butane.

In Malaysia, the LPG compositions and specifications are obtained from the Energy Commission (Suruhanjaya Tenaga). The composition of LPG in Malaysia, as shown in Figure 1 below, consists of 3 main compounds, propane, isobutane and n-butane at a ratio of 403030 (Suruhanjaya Tenaga, 2008). Other properties of LPG in Malaysia are summarized in Table 2.

Gas	Mol (%)
C_3H_8	40.0
iC_4H_{10}	30.0
nC_4H_{10}	30.0

Figure 1: Typical composition of LPG in Malaysia

Table 2: Typical LPG properties

Specific Gravity	1.65
Gross Calorific value (kcal/Sm3)	28 059
Burning Velocity (m/s)	0.46
Upper Flammability Limit (%)	8.5
Lower Flammability Limit (%)	1.9
Auto ignition Temperature (degrees Celcius)	510
Theoretical Air Requirement (m3/m3)	28.81

LPG is colourless and odourless. Due to that, odorant is added in the bottled LPG so that leaks can be detected. When compared relatively with air, LPG is 2 times heavier.

1.1.2 Normal Butane

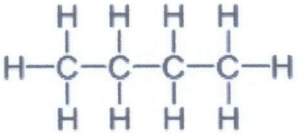

Figure 2: N-butane molecular diagram

Figure 3: N-butane skeletal diagram

Normal butane or n-butane is one of the gaseous components of natural gas and also a component produced from crude oil, but in smaller quantities (Pollick, 2013). It has a molecular formula of C_4H_{10} and a molecular weight of 58.12 g/mol. At room temperature and atmospheric pressure, it exists in the gaseous state because of its boiling point which is from -1°C to 1°C (272 to 274 Kelvin). It is colourless and has a density of 2.48 g/dm^3 (at 15°C). Mostly, n-butane is non-toxic and has odour similar to natural gas or petrol.

Its properties of being highly flammable and easy to be liquefied, makes it suitable to be used as fuels. As mentioned previously, butane is mixed with propane to form LPG. Another main uses of n-butane is that it is used as raw material in the production of 1, 3-butadiene, ethylene and acetic acid (Vesovic, 2011). Butadiene which is produced from n-butane can be used to manufacture synthetic rubber and polymerization for plastics manufacturing (SABIC, 2013). Other use of n-butane includes as a gas for refrigerant and as solvents. As for ethylene, it can be used for various applications such as in producing polyethylene, as raw materials in manufacturing surfactants, detergents and also plastics used for soft drinks (Shell, 2008). Properties of n-butane are summarized as shown in Table 3.

Table 3: Properties of n-butane

Properties of N-butane	
Molecular weight	**58.12**
Melting point	**134.82K**
Normal boiling point	**272.66K**
Normal vapour density	**2.59 kg/m3**
Critical temperature	**425.16K**
Critical pressure	**3.796MPa**
Critical density	**225.3kg/m^3**

1.1.3 Isobutane

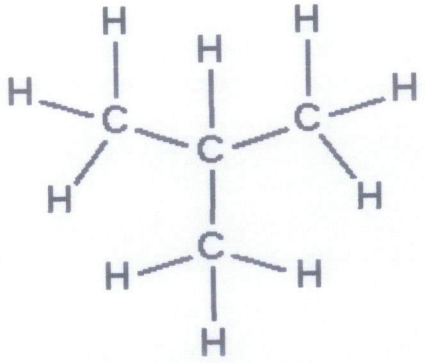

Figure 4: Isobutane's molecular diagram

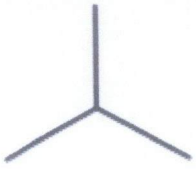

Figure 5: I-butane's skeletal diagram

Isobutane is also known as i-butane or methylpropane. Although i-butane has the same formula with n-butane, it is different because it has different arrangement of atoms in the molecule and different properties as well; in short, it is an isomer of butane. Similar to n-butane, it has molecular weight of 58.12 g/mol. One of the most common uses of isobutane is for the manufacturing of aerosols sprays, commonly

referred as R-600a. The addition of isobutane to the spray gives propellant qualities (Tatum, 2013). Freon which has been used as a compound that helps refrigeration is now being replaced with isobutane because isobutane does not pose any harm to the environment as compared to Freon gas which would damage the ozone layer.

Air-conditioning is another application which uses isobutane thus, posing huge potential in the refrigeration field. Isobutane is alleged to release lesser greenhouse gases (GHG) but it was also claimed to be more efficient refrigerant than chlorofluorocarbons (CFC) (Traum, 2009). Another advantage of isobutane is that it has zero ozone depletion potential (ODP) which means that it does not cause any harm to the ozone layer. Despite of its advantages, isobutane which is flammable and at very high concentration it is toxic to humans, are the key factors they are yet to be utilized widely (Traum, 2009). Extremely high concentration of isobutane may cause cardiac effects and narcosis (Hathaway & Proctor, 2004). In the future, isobutane has huge prospective to replace gases such as chlorofluorocarbon or hydrofluorocarbon in air-conditioning systems.

In the petrochemical industry, isobutane is used as feedstock for isooctane production because isobutane aids in creating the ideal environment for such production (Tech-FAQ, 2012). The purpose of isobutane is to produce refinery alkylate which will be blended with motor gasoline to reduce the "knocking" effect of engine combustion (Matchell, 2012). Higher amount of isobutane is used for blending with gasoline during the cold season to ensure that the gasoline used does not freeze. The purpose of the blending of gasoline with isobutane is because of isobutane's low melting point that is -159.6°C. This would result in the mixture having lower melting point thus ensuring that the fuel does not freeze especially during winter.

1.1.4 Economic Potential and Price Trending

Figure 6 illustrates the prices of natural gas liquids in 2012. Both, normal butane and isobutane price trend follows the price trend of natural gas liquids (NGL) and crude oil mainly because they are cracked or refined from them. Aside from that, the demand for the isobutane and normal butane is another factor which will affect the price of normal butane and isobutane.

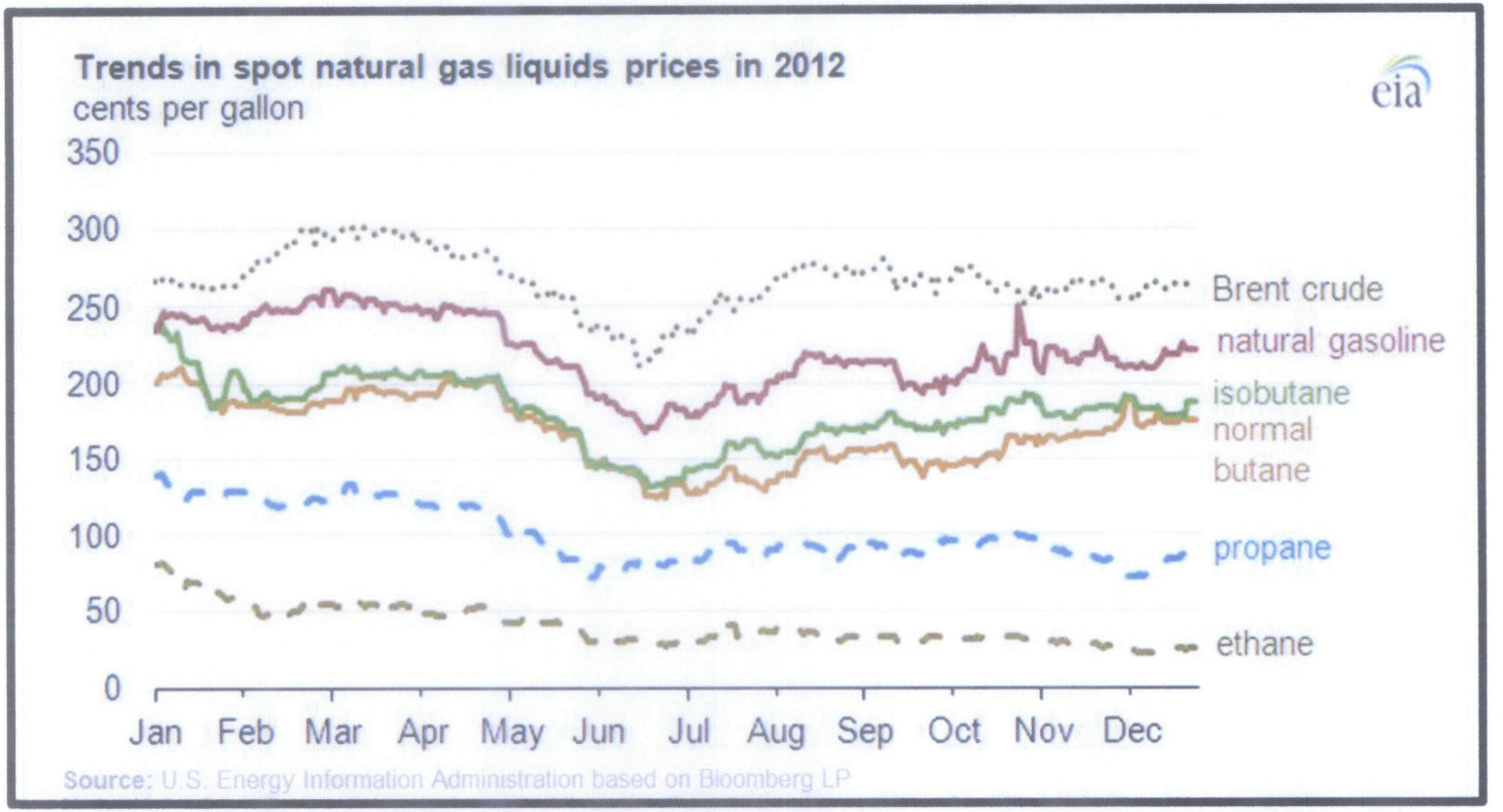

Figure 6: Natural Gas liquids prices in 2012

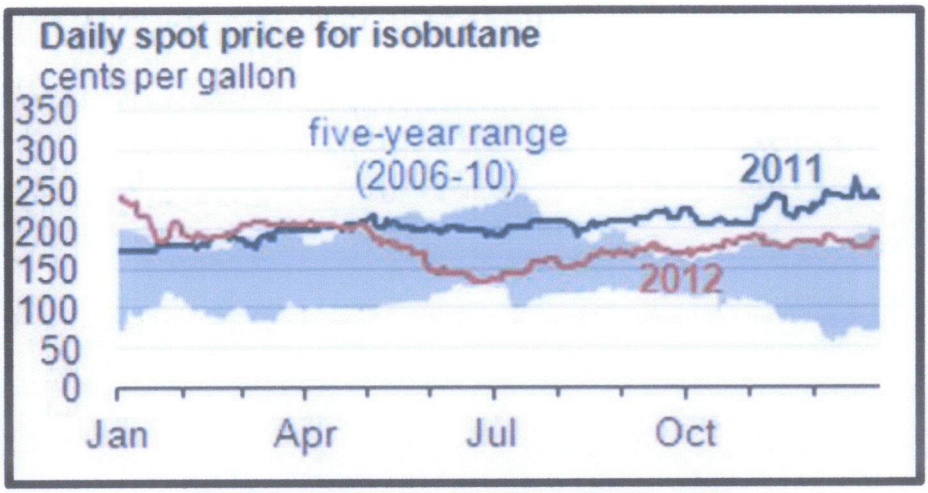

Figure 7: Isobutane price from 2006 – 2012

Figure 7 and Figure 8 show the trend of isobutane and normal butane respectively. The price of normal butane which is traded around 1.10 to 1.20 USD per gallon. As for isobutane, the price is often slightly higher than normal butane (Leffler, 2000). LPG which consists mainly of propane is traded at about 0.84 USD per gallon. This shows that the C_4 hydrocarbon mixtures in LPG, isobutane and normal butane, have the potential added value if they were extracted out from the LPG gas supplied by PETRONAS.

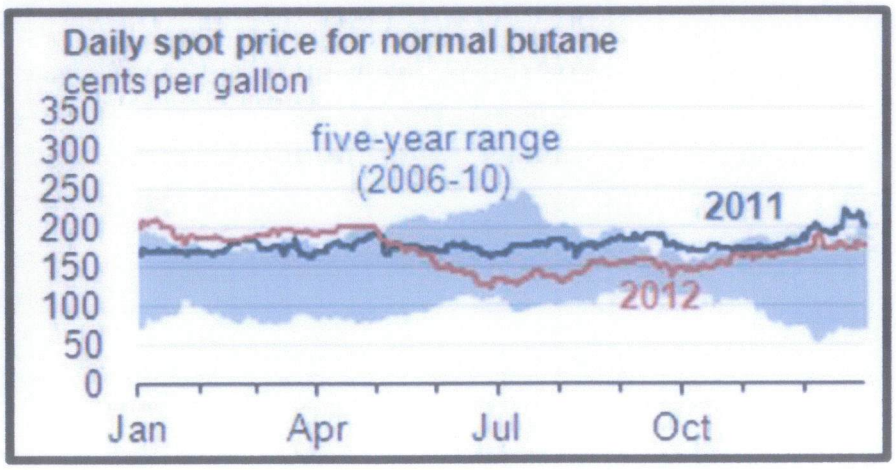

Figure 8: Normal butane price from 2006 - 2012

1.2 Distillation

Distillation is an important separation process that is used to separate a mixture of components into their pure products. Its application in numerous industries has made the subject to be researched upon for improvements in terms of energy utilization, operating costs reduction and flexible distillation design (Bennett & Kovak, 2000). Typical design optimization variables include pressure, total number of trays, feed tray location and sidestream tray location (Luyben W. L., 2011). These variables are important because they will affect conversion, selectivity, capital investment and operating cost. Distillation column is one of the most common operations in process industries and uses the most amount of energy consumption (Masoumi & Kadkhodaie, 2012).

The normal operating conditions for the feed should be in liquid state and minimize vapour state flow rate for a distillation column (Luyben W. L., 2013). This would reduce loading in the distillation column. The split between normal butane and isobutane through the conventional distillation of normal butane and isobutane is difficult because they have close relative volatilities. Thermodynamic information that is widely available, the mean relative volatility of normal butane and isobutane is about 1.20, under the desired operating temperature and pressure.

$$Nmin = \frac{Log\left[\left(\frac{d}{b}\right)_{LK}\left(\frac{b}{d}\right)_{HK}\right]}{log\ \propto_{LK,HK}}$$

Equation 1: Fenske Equation

Where

N_{min} = minimum number of equilibrium stages
$\propto_{LK,HK}$ = relative volatility between the LK and HK components
d, b = distillate and bottoms molar recovery fractions, respectively

In order to obtain the basis on the number of stages in order for simulation to be done, the equation could be used. Fanske equation can be used to calculate the minimum number of equilibrium stages, N_{min}, for a separation given the light key

(LK) and heavy key (KH) components (Quantrille & Liu, 1991). Due to the relative volatility of isobutane and normal butane that is small, the number of stages of butane splitter can be expected to be higher than depropanizer (the distillation column to separate propane from the butane mixtures) thus, increasing overall costs. (Rafiq, Maros, Majed, Far, & Muhammad, 2011).

As for the minimum reflux ratio (R_{min}), estimation can be done using the Underwood equations as shown in the Figure 9.

$$\sum_{i=1}^{n} \frac{\alpha_i x_{i,F}}{\alpha_i - \theta} = 1 - q$$

$$R_{\min} + 1 = \sum_{i=1}^{n} \frac{\alpha_i x_{i,D}}{\alpha_i - \theta}$$

Figure 9: Underwood equation

Where n, number of components
q, liquid fraction of feed
$x_{i,F}$, mole amount fraction of I in feed
$x_{i,D}$, mole amount fraction of I in distillate
α_i, relative volatility of component I to the heaviest component
θ, root of equation that can be found from the first Underwood Equation

For the actual number of stages (N), the Gilliland correlations can be applied. The equation is shown in the following Figure 10.

$$Y = 1 - \exp\left[\frac{1 + 54.4X}{11 + 117.2X} \cdot \frac{X - 1}{X^{0.5}}\right]$$

where,

$$Y = \frac{N - N_{\min}}{N + 1}; X = \frac{R - R_{\min}}{R + 1}$$

R, actual reflux ratio.

Figure 10: Gilliland correlations

The feed tray location can be identified using the Kirkbride equation.

$$\log \frac{N_r}{N_s} = 0.206 \log \left[\frac{z_H}{z_L} \frac{F_B}{F_D} \left(\frac{x_{B,L}}{x_{D,H}} \right)^2 \right]$$

where,
N_r, number of stages above feed stage
N_s, number of stages below feed stage
z_H, mole amount fraction of heavy key component in feed
z_L, mole amount fraction of light key component in feed
F_B, molar flowrate of bottom product
F_D, molar flowrate of distillate product
$x_{B,L}$, mole fraction of light key component in bottom product
$x_{D,H}$, mole fraction of heavy key component in distillate product.

Figure 11: Kirkbird equation

1.3 Problem Statement

From an enterprise perspective, the separation of isobutane and normal butane from the LPG supplied by PETRONAS from the LPG bottled has the potential for added value creation. The current usage of the LPG which is used as fuels contains propane, isobutane and normal butane. They can be separated and be sold as individual product that has potential to add value because the market price of isobutane and normal butane is higher than LPG. The price of propane, isobutane and butane are RM 2.86/gal, RM 4.64/gal and RM 4.47/gal respectively. The price of LPG in Malaysia is at RM 1.90 per kg and they are fixed due to the regulated price by the Malaysian Government. Thus, the feasibility of the project depends on the sales margin that could be maximized from the amount of volumes that could be sold against the operating cost of separating them.

The conventional distillation method which requires large number of trays and large reflux ratio are mainly due to the close relative volatility of normal butane

and isobutane. "Achieving higher concentration of isobutane requires more stages and thus increasing capital costs" (Luyben W. L., 2011). Aside from that, using distillation alone, it is almost impossible to obtain high purity isobutane unless olefins in the mixture is hydrogenated with catalytic reaction then separated using distillation (Han, Park, Kim, & Cho, 2003).

To summarize, the main problem statement of this study is as follows

- Separation of LPG requires high energy and large number of trays for the separation of isobutane and normal butane due to the close relative volatility
- Could the alternative separation processes provide optimization in terms of energy reduction in the separation of LPG into its constituents?

1.4 Objectives

The current practice in Malaysia which bottles LPG mixtures containing n-butane and isobutane for domestic and commercial use has additional potential for further separation involving propane, isobutane and normal butane. Thus, this study is aimed at

a) To develop a base case simulation involving the separation of propane, normal butane and isobutane
b) To identify the amount of energy utilization
c) To simulate alternative separation processes to study on potential optimization in terms of energy utilization

1.5 Scope of Study

The scope of the study will focus on the separation of propane, isobutane and normal butane that originates from the bottled LPG in Malaysia which constitutes of the 3 components. It is assume that other components such as ethane and pentane that exists in LPG in trace amounts are negligible. Their individual functions especially isobutane as chemical feedstock for the production of other compounds raises the possibility for value creation of LPG. A simulation of the base case of the conventional distillation will be done using Aspen HYSYS. From the base case developed, potential optimization areas especially in terms of energy utilization will be studied and consequently, simulated in the software, Aspen HYSYS.

1.6 Relevancy of the Project

The project is relevant because it would require significant amount of chemical engineering technical competency and knowledge. The need of equipping with some fundamental understanding in the downstream of oil and gas is crucial in this project because of the mixtures of compounds involved such as liquefied petroleum gas (LPG), propane and butane. Aside from that, in the later part, the separation of the hydrocarbons will require much background and in depth study on the technical part thus, connecting back to the chemical engineering knowledge. Distillation is the key study in this project which will require in depth analysis in order to achieve the objective of this study.

1.7 Feasibility of the Project

The project involves 2 semesters which begun on May 2013 and is expected to end on December 2013, which gives the project about 28 weeks to be completed. Divided into 2 semesters, namely Final Year Project I (FYP1) and Final Year Project II (FYP2), the first part included identifying the motivation of the study, problem

identification and up to the initial part of the project execution which comprised of the simulation of the base case study using Aspen HYSYS. It also comprised of proper planning of the project and ensuring that the time frame provided was sufficient. It is important to ensure that the project feasible within the time frame and feasible to be conducted within the authors' ability so that the project could produce results that at the end of the timeline.

As for the second part, Final Year Project 2 (FYP2), it will encompass execution of the optimization simulation and followed by the completion of a full report and the presentation of the results to the panel of examiners- internal and external panels. The author will be evaluated in several means such as oral presentation and through the dissertation produced by the author. It is therefore, important that the project be planned and executed in a timely manner to ensure that within the time frame provided, the author will be able to produce a good study and at the same time be able to generate successful results.

In terms of the study or project's perspective, the preliminary analysis on the feasibility showed that the project has potential for value creation. The price of LPG which is cheaper than the products, isobutane and normal butane demonstrated that by extracting them from the LPG, they can be utilized for various purposes that have higher value. However, added value gained from the products will have to be weighed against the total costs of processing or separating the desired products. If the process, in this case, in terms of energy can be optimized and the profit margin is enough to generate profit, this study can further be scrutinized and conducted in depth to study the feasibility of setting up the processing plant.

CHAPTER 2

LITERATURE REVIEW

2.1 Sequencing of Distillation Column

Masoumi and Kadkhodaie (2012) scrutinized on the heat integration of the distillation column. The study involves the separation of benzene, toluene, xylene and C9+ components. The study investigates in 3 cases namely, conventional heat integrated distillation, heat integrated distillation with forward energy integration and heat integrated distillation with backward energy integration. Conventional heat integrated distillation is as shown in Figure 12. This distillation uses the energy supplied by the condenser and reboiler of the columns. In Figure 12, cooling water (CW) and steam is used as the medium to supply the energy required. In heat integrated distillation with forward energy integration as shown in Figure 13, heat from the vapour from the first column is utilized in the second column to reboil the liquid at the bottom of the column. This reduces the amount of utility because in the second column, if energy supplied is sufficient, the reboiler unit may not be required.

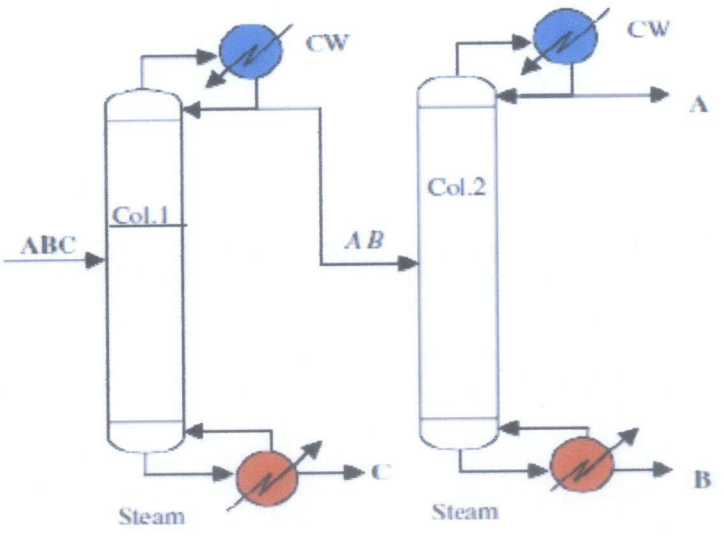

Figure 12: Conventional heat integrated distillation

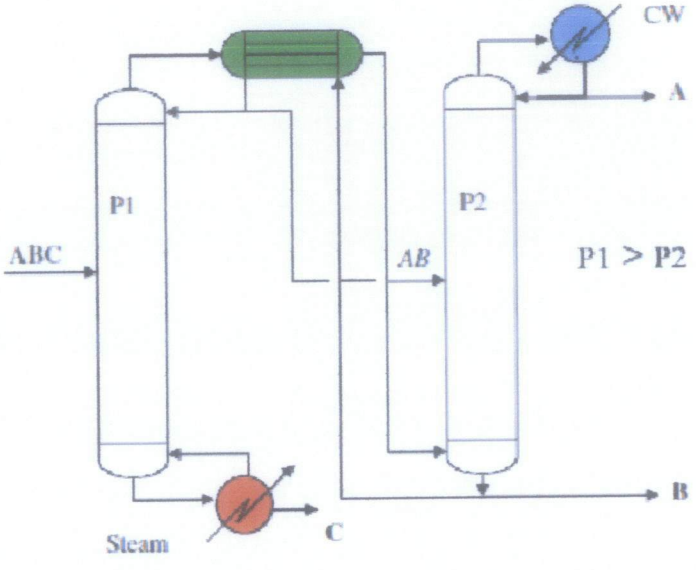

Figure 13: Heat integrated distillation with forward energy integration

In Figure 14, the heat integrated distillation with backward energy integration is illustrated. In this setup, opposite to the process shown in Figure 13, heat from the second column is utilized in the first column to reboil the liquid at the bottom. Stream from the top of the second column goes through a heat exchanger to supply energy to the bottom of the first column. It then, goes back to the top of the second

column to serve as a medium for the condenser because the stream now, has lower energy due to heat transfer in the heat exchanger.

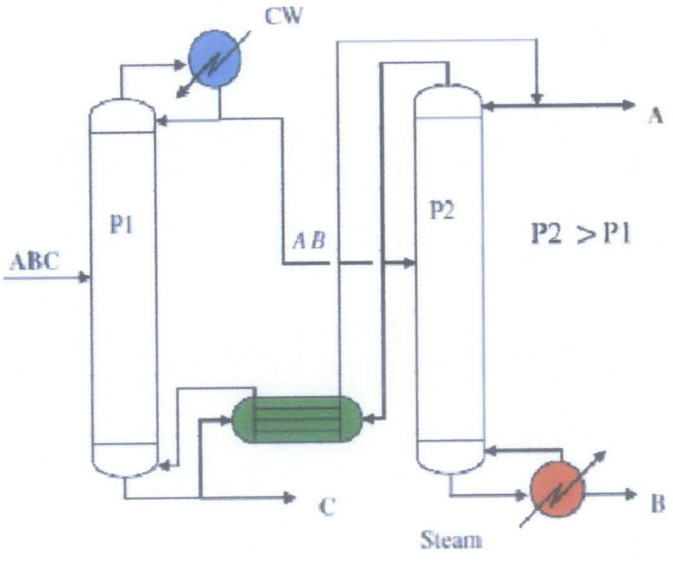

Figure 14: Heat integrated distillation with backward energy integration

The study has shown that the energy consumption was reduced by 33% for heat integrated distillation with forward energy integration as compared to the conventional heat integrated distillation. As for heat integrated distillation with backward energy integration, the study stated a reduction of 28% in terms of energy consumption as compared to the conventional heat integrated distillation. Subsequently, reduction in energy consumption will also affect the costs and economics of the distillation operations. In this case, the study calculated that annually, cost reduction of 12% can be obtained for forward heat integration and 8% for backward heat integration (Masoumi & Kadkhodaie, 2012).

Another study by Harwardt, Kossack and Marquardt (2008) involves a study on the sequencing of distillation column to separate of zeotropic mixtures. The mixture involves pentane, hexane, heptane and octane. The study assumed a pressure of 1.013 bars with equimolar feed and total flowrate of 10 mol/s. The case study

involved 5 different distillation sequences as shown in Table 4. However, the breakdown of the minimum energy was not indicated. Feed composition is important in determining the optimal sequencing of the distillation column (Harwardt, Kossack, & Marquardt, 2008). The simulation of each sequence gives the optimal sequence from the one that requires the lowest energy. The separation indicated that different separation whether direct or indirect would result in significant difference in the energy requirement for separation. Thus, in this project, the base case will also be the direct sequence and will be compared with the energy utilized for indirect sequences.

Table 4: Task and sequence energy

Task #	Feed	Top	Bottom	Min. Energy (kW)
1	ABCD	ABC	D	366
2	ABCD	ABC	CD	602
3	ABCD	ABC	BCD	242
4	ABCD	ABC	CD	595
5	ABCD	ABC	DC	257
6	ABCD	ABC	B	311
7	BCD	BC	D	329
8	BCD	BC	CD	465
9	BCD	BC	CD	476
10	CD	C	D	193

Note Acetone (A), acetone/chloroform (B), benzene (C) and toluene (D)

In a separate literature, Jain et al. (2012) works on the energy minimisation through inter-column heat recovery. A reboiler of a column will exchange heat with a condenser of another column however the minimum temperature approach has to be obeyed. It could be done through manipulating the pressure of the columns (Jain, Smith, & Kim, 2012). The study has shown that the heat integration can be optimized thus reducing the costs of operating the distillation columns. Another method to optimize energy is through the introduction of heat pump whereby heat from a lower temperature source is upgraded to a higher temperature source (Bruinsma & Spoelstra, 2010). This approach uses a compressor that increases the pressure and temperature, which will then be used in the reboiler for evaporation.

As for vapour compression (VC), it uses working fluid evaporated at the condenser and compresses it to a higher temperature before feeding it into the reboiler, which will then condense and cool down due to expansion (Bruinsma & Spoelstra, 2010). In vapour recompression (VRC), the mechanism is almost similar whereby the fluid at the top of the column is compressed and condensed in the reboiler. The main difference is that the working fluid will then be partially refluxed to the top of the column after its pressure has been reduced. Comparing VC and VRC, VRC has higher thermodynamic efficiency because heat is exchanged once and the condenser in VRC is smaller. Figure 15 shows the pressure and temperature graphon the modelling developed in the literature. It shows the optimized temperature difference over the heat exchanger, ΔT_{HEX}. It was assumed that the trim-condenser is used to de-superheat the compressed vapour denoted as E.

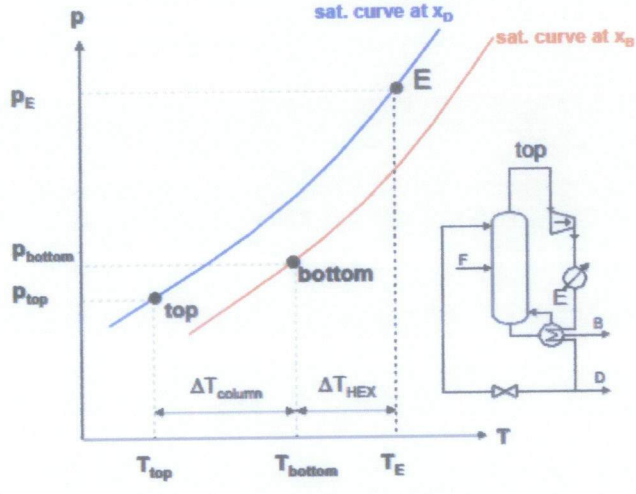

Figure 15: PT diagrams of VRC cycle

2.2 Dividing Wall Column (DWC)

The conventional distillation columns are simple to control and operate. However, they can be inefficient because of the occurrence of mixing entropy by irreversible split (Long & Lee, 2011). If thermodynamic efficiency is improved (reducing remixing and re-separation) the amount of utilities used can also be reduced (Manley, 2000). Another study by Jung (2012) also states that the dividing

wall column (DWC) and bottom dividing wall column (BDWC) reduces energy consumption and thus leading to reduction in operating costs.

Dividing wall column allows the reversible splits to be performed instead of having an external prefractionator. DWC has a single shell that consists of an internal wall that acts as a prefractionator. The internal wall which divides the columns is also known as dividing wall column. According to literatures, the DWC does the separation of reversible splits once as compared to conventional columns which do it twice (Jung, Long, Woldetensay, & Lee, 2012).

DWC which requires only a single column thus, it provides added advantage of energy savings as it requires less capital expenditure and space. Estimated to reduce capital expenditure up to 30% as compared to a conventional 2-columns sequences, the design of the DWC is complex because higher number degrees of freedom (DOF) that needs to be specified such as the recycle feed inlet stage, composition of the recycle streams and etc. (Jung, Long, Woldetensay, & Lee, 2012). Another literature stated that separation of 3 components feed into 3 products is difficult as compared to the conventional 2 columns configuration because they can be optimized independently of each other (Matijesevic & Olujic, 2010). Figure 16 illustrates a typical DWC with the division configured in the middle.

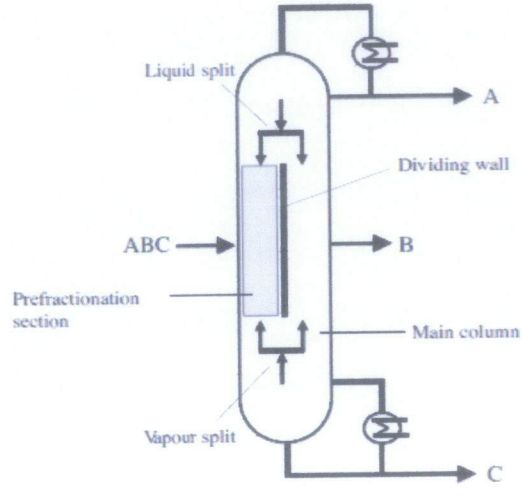

Figure 16: Dividing Wall Column

Dividing wall column can further be broken down to several types however in this case the wall of the column can be configured to be in the middle, lower or the upper part of the column as shown in the Figure 17.

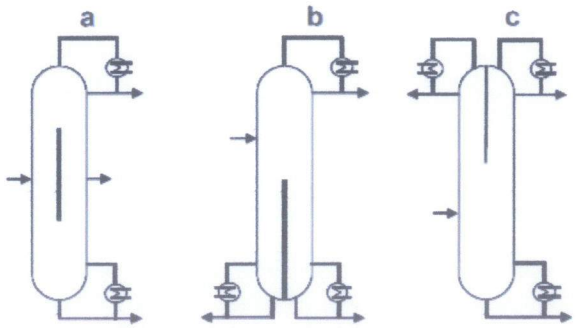

Figure 17: Basic types of DWC

Another article by Yildirim et al. (2011) stated that DWC can reduce energy consumption and consequently reduce the operating costs however they are more complicated to be controlled. This is due to the, additional degree of freedom of the liquid flow in the divided wall section (Yildirim, Kiss, & Kenig, 2011). As for the pressure of DWC, in conventional columns, their pressure can be specified individually however, in DWC, only one pressure specification is possible. Another limitation of DWC is the height requirement of DWC increases as compared to individual distillation column (Asprion & Kaibel, 2010). This will result in a higher capital expenditure or investment.

As for petluk arrangement as shown in Figure 18, it is thermally coupled where the column has only one condenser and one reboiler that are in the second column as shown in the Figure 18. Vapour and liquid from the second column are used to provide liquid for the prefractionator. According to Schultz et al. (2002), this arrangement may reduce capital costs because it has fewer pieces of major equipment as compared to the conventional two-column sequence. The study also stated that incorporating the prefractionation column into the main column will further reduce the capital costs by 30%. Theoretically, the Petlyuk arrangement is suitable for distillation separation where 2 products are to be separated however in reality, much deliberation and study has to be done because often parameters such as

temperature, pressure, height, liquid and vapour load can be very different between sections.

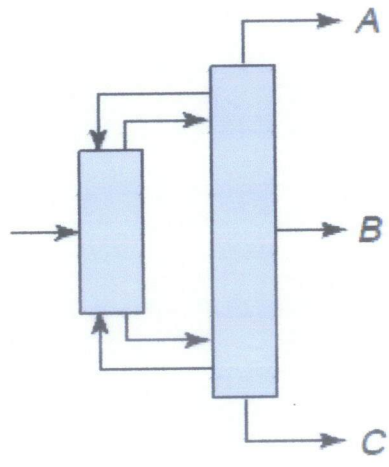

Figure 18: Petyluk Column

2.3 De-isobutanizer Plant Study

Figure 19 illustrates a study at increasing the throughput of the de-isobutanizer unit through modification of equipment. Also, in the study, results shown that the throughput could have been doubled. The change of trays to high performance trays in an existing plant was found to be cost effective and the production capacity could be stepped up (Mosca, Hemel, & Tocco, 2005). The increase in the production of i-butane required the existing equipment to be insufficient in the process thus they had to be modified to cater to the increase of capacity and loadings. Figure 19 highlights the modifications done on the existing unit to increase its efficiency.

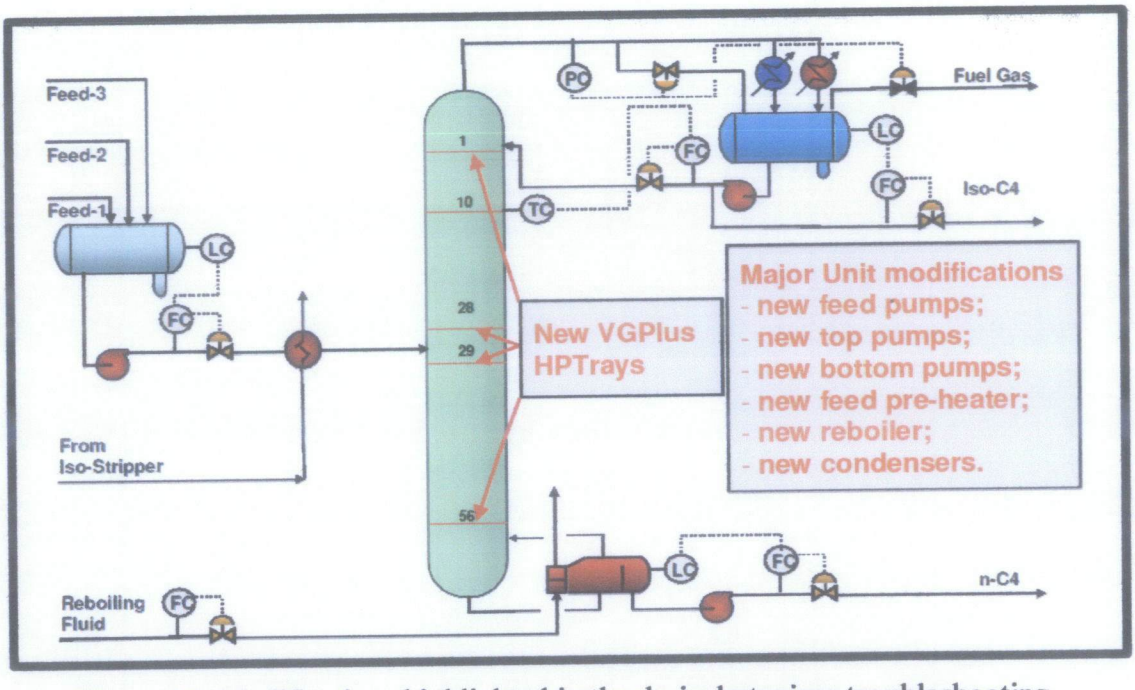

Figure 19: Modifications highlighted in the de-isobutanizer troubleshooting (Mosca G. , Hemel, Tocca, & Lestrade, 2006)

Another research paper by Klemola and Ilme (1996) simulated an industrial scale of the isobutane and normal butane fractionator which contains propane, butane and pentane. The specifications of the column and the performance data are shown in Table 5.

Table 5: Column specification and performance data
(Klemola & Ilme, 1996)

Column specification and performance data			
Feed tray			37
Feed flow rate, kg/h			26234
Bottom flow rate, kg/h			18119
Distillate flow rate, kg.h			8115
Reflux flow rate, kg/h			92838
Reflux temperature, oC			18.5
Column top pressure, kPa			658.6
Feed pressure, kPa			892.67
Boiler duty, MW			10.24
	Composition, wt %		
	Feed	**Top**	**Bottoms**
Propane	1.5	5.3	0.3
i-butane	29.4	93.5	0.3
n-butane	67.7	0.2	98.1
1-butene	0.2	0.4	0.1
i-butene	0.2	0	0
trans-2-butene	0.1	0.6	0.1
neopentane	0.1	0	0.2
i-pentane	0.8	0	1.1
n-pentane	0.1	0	0.1

From the literature, the research discovered that the ideal number of trays was found to be 88 and the actual number of trays to be 74 (Klemola & Ilme, 1996). The stream for the column was at the 37th tray. This study provides basis for the butane splitter which can be used in the base case simulation of this study since the composition of the research paper even though the components varies by having mixtures of hydrocarbons from C_3 to C_5 as compared to LPG which is only contains C_3 and C_4 hydrocarbons.

2.4 Butane Splitter

In a petrochemical or oil and gas fractionation plant, the butane splitters which will be used to separate isobutane from the heavier normal butane requires high number of trays because they have very similar volatilities. Moreover, high reflux ratio is needed. The typical values of reflux ratio of a depropanizer and butane splitter column are as shown in Table 6.

Table 6: Typical and Recommended Efficiency of Fractionation Columns (Cetac-West, 2008)

Column	Reflux Ratio
Depropanizer	1.8 – 3.5
Butane Splitter	6.0 – 14.0

The case provided an initial estimate for the reflux ratio needed in the column. In the fractionation plant study done by Jung et. al., (2012), it involves the separation of propane, iso-butane, normal butane and pentane plus. The study has selected pressure of about 14.00 bars for the depropanizer and 5.50 bars for the butane splitter (Jung, Long, Woldetensay, & Lee, 2012). Using this information, an initial estimate for the simulation can be conducted as the components are close in terms of properties for this study.

In spite of the said similarities, the literature contains more components in its streams as compared to the case in this study. A typical fractionation plant contains more components ranging from light hydrocarbons to heavier ones. However, as mentioned previously, in this study, only propane, normal butane and isobutane will be simulated as other components are assumed to be in trace amounts thus, negligible. Therefore, some parameter such as pressure requirement may not need to be as high. Operating the plant at high pressure will impact the costs because high pressure materials will be required for the construction of the plant, making it less economical.

Typical selection of operating pressure of a distillation column depends on the cooling medium (cooling water) that is cheap and able to condense the distillate vapour (Thakore & Bhatt, 2007). Thus, the bubble point of distillate should be higher than the cooling medium for condensation to occur. In order to cater for the conditions, the streams will be pressurized to an operating pressure based on the cooling medium (in this case, cooling water). A high operating pressure can further be reduced by replacing the cooling water with chilled water or brine.

CHAPTER 3

METHODOLOGY

3.1 Research Methodology

Much of the study involves execution using process simulations using engineering software namely, Aspen HYSYS. After conducting research and understanding the subject matter, using the software, the simulation of separation of C_4 hydrocarbon components would be done. The LPG compositions for this study utilized was taken from Gas Malaysia specifications. Additional work such as process optimization will also be conducted so that the processes would be more feasible and economical. A process flow diagram regarding expected work flow is as follows

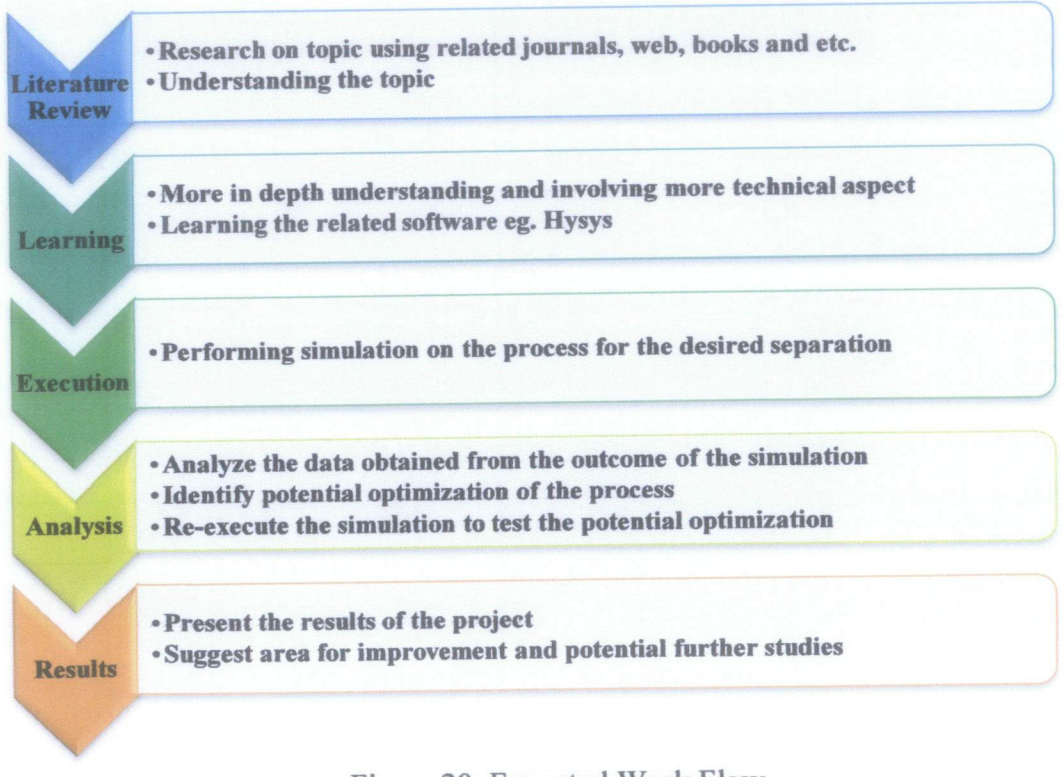

Figure 20: Expected Work Flow

3.2 Gantt Chart and Key Milestone for FYP

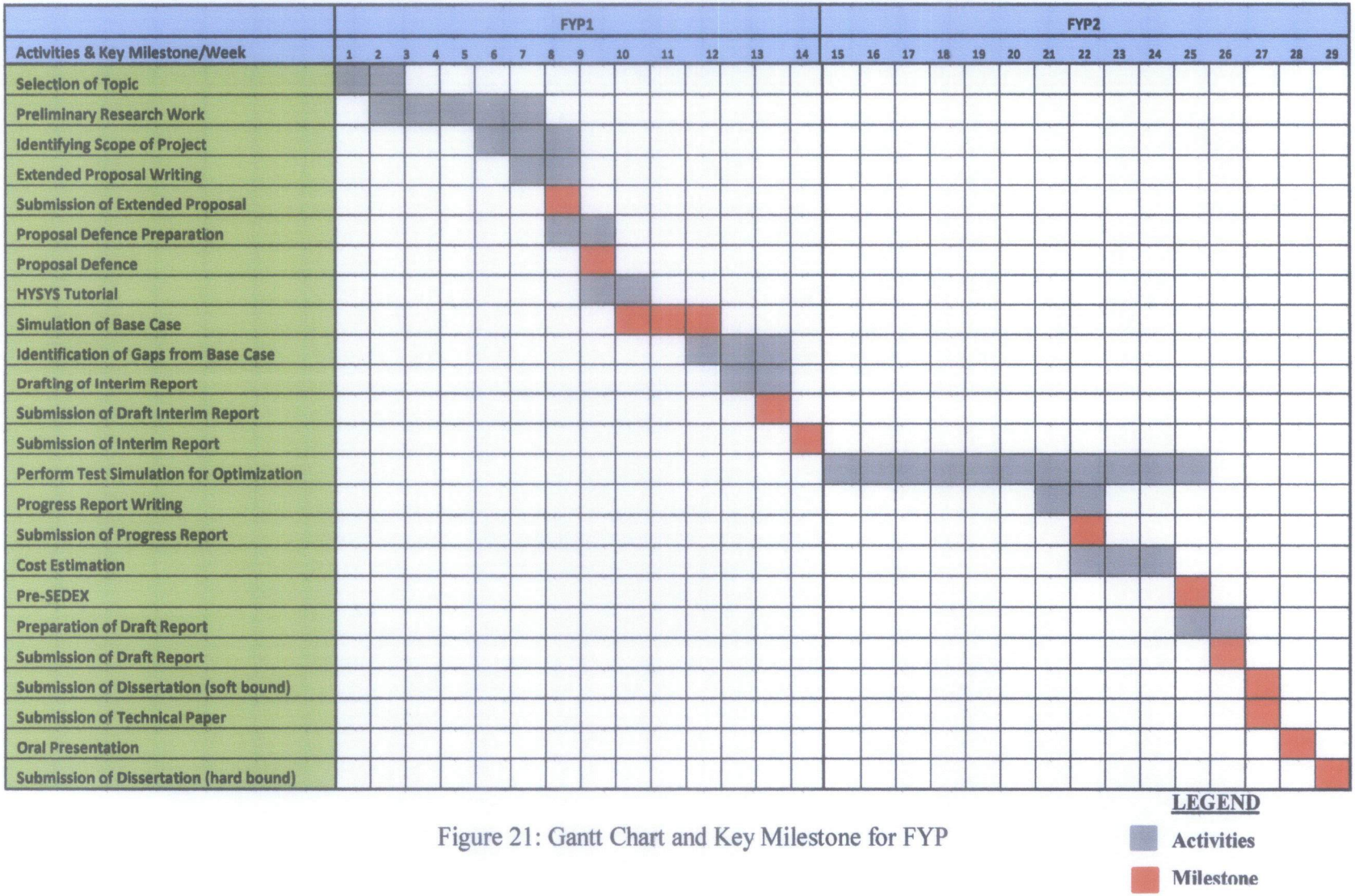

Figure 21: Gantt Chart and Key Milestone for FYP

3.3 Tools, Software and Equipment

The final year project in this study utilizes Aspen HYSYS software. The version of Aspen HYSYS used in this study was Version 2006 (20.0.0.6728). This simulation tool can integrate dynamic simulation for oil and gas, gas processing, petroleum refining and other industry processes. Aspen HYSYS is claimed to be a comprehensive process modelling system used for oil and gas, refineries and engineering companies to optimize process design and operations (AspenTech, 2013). The detailed methodology of this tool will be discussed in the next section of this chapter.

3.3 Simulation of Base Case System (Direct Sequence)

The base case was established in HYSYS after establishing some related parameters based on literatures of the separation of propane, normal butane and isobutane. The conventional distillation column used as the base case is typical and is widely applied in the industries. Since the composition of the LPG only constitutes propane, normal butane and isobutane, they were selected as the components in HYSYS, as shown in Figure 22. Traces amounts of ethane and pentane is assumed to be negligible and is therefore disregarded in the study. The mole fraction of each component was specified in HYSYS as shown in Figure 22 below based on Gas Malaysia specification.

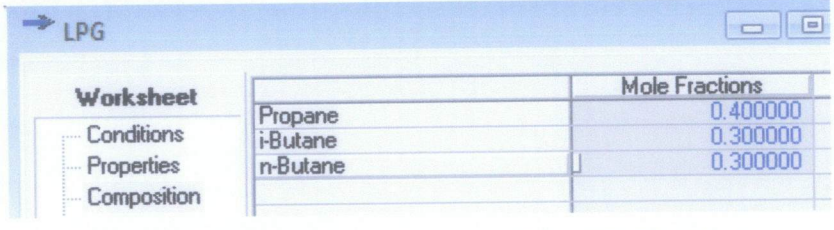

Figure 22: LPG Feed Composition Specified in Hysys

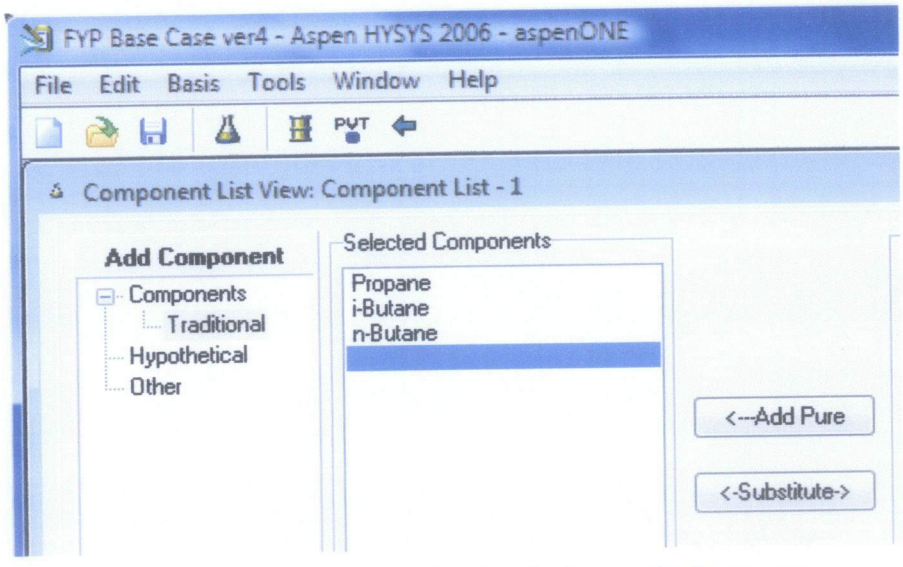

Figure 23: Components selection in hysys for base case

Peng-Robinson property packages or simulation basis manager, as shown in Figure 23, was chosen. Simulation basis manager or property package is important in the simulation as the settings selected will be effecting the thermodynamics interaction of the simulation. In this case, literatures indicated that the Peng-Robinson can represent the thermodynamics interaction for this study's simulation. According to Prausnitz et al., Peng-Robinson is common and useful for calculation which involves oil, gas and petrochemical industries (Prausnitz & Tavares, 2004). Another source also stated that when dealing with hydrocarbons, Aspen Peng-Robinson equation-of-state property package is chosen because it accurately describes the thermophysical properties of hydrocarbons mixture (Folger & Gurman, 2007). Aside from that, Peng-Robinson equation of state can calculate accurately for liquid and vapour phase (Binous, 2008). The following Figure 24 shows the selection of Peng-Robinson Property Package in HYSYS.

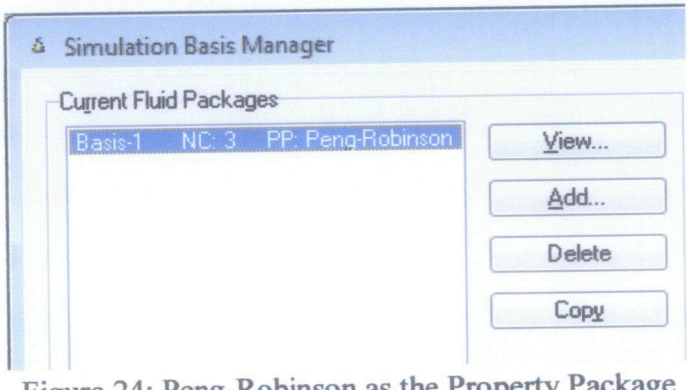

Figure 24: Peng-Robinson as the Property Package

The process was set up and distillation column was made to converge successfully. Parameters and conditions were manipulated after the base case converged. For the base case which is the direct sequence configuration, the LPG inlet pressure that was feed to the first distillation column was varied between 700 kPa to 1500 kPa for the Distillation Column 1 (DC1). Figure 25 is a snapshot of the base case simulation for the DC1. In the base case simulation, a pressure of 1,400 kPa was used because this will raise the temperature of the exit streams to a higher temperature so that water can be used as the cooling medium in the condenser. The downside of having a high pressure is that the construction material will need to withstand higher pressure thus they will be more expensive and maintenance costs would also increase.

Stream that has low temperature will require chilled water instead of cooling water for cooling purposes which at best, should be avoided. The temperature of chilled water is in the range of 4 to 7°C whereas for the cooling water, it is in the range of the ambient temperature which hovers around 30°C. Cooling water is the preferred choice as compared to chilled water because it is cheaper than chilled water. As for the reflux ratio, the base case specification was simulated as recommended by a literature that was from 1.8 to 3.5 for DC1. The pressure drop for the column was kept constant at 30 kPa for Distillation Column 1 (DC1) and Distillation Column 2 (DC2). Assumption of no energy losses in the system were implied for the processes for example the heat energy losses to the surroundings are not significant and they have been accounted for by the simulation software.

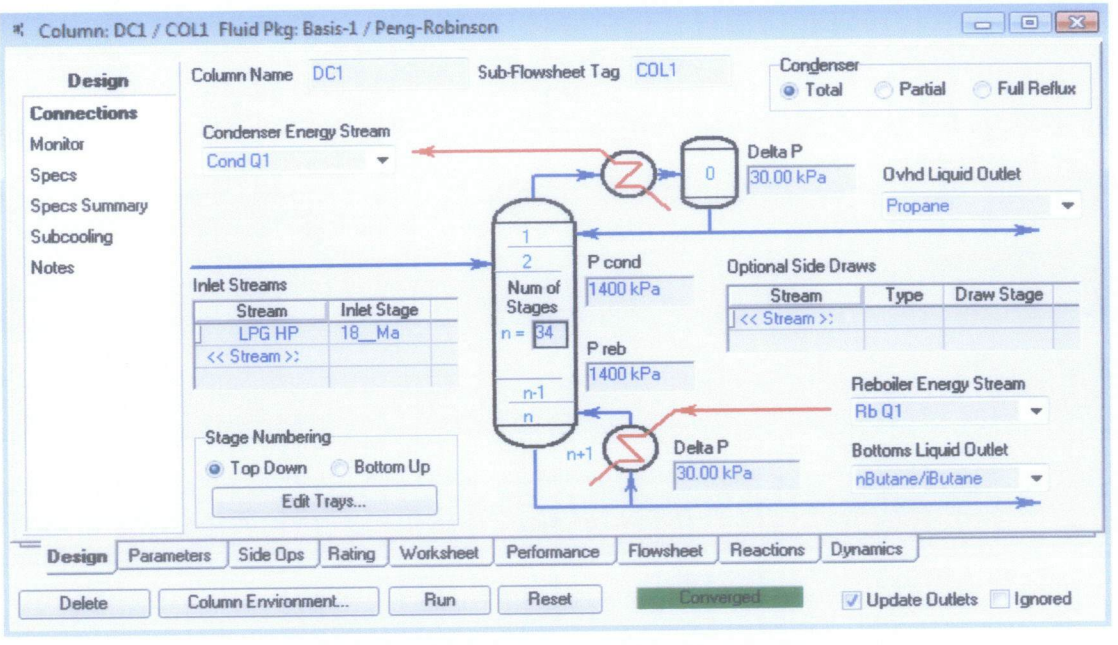

Figure 25: Base Case Simulation for DC1

The inlet stream for the DC1 enters at the 18th stage. The reflux ratio used was 3.5. The total number of stages that was simulated for DC1 was 34 stages, in order to achieve the desired separation. The propane product purity obtained from the DC1 was 90% of mole fraction. As for DC2 (refer to Figure 26 for the simulation snapshot from HYSYS), the reflux ratio was specified to be 10.07 and the column was simulated with 74 stages and its inlet stream fed at 37th stage. The iso-butane product obtained was 97% mole fraction. As for the normal butane, 97.5% of mole fraction was obtained. The summary of the purity of propane, isobutane and normal butane is shown in Table 7. This purity will be kept constant for all the cases as the variable that is kept constant in this study. At the specification shown Table 7, the purified constituents can be utilized as a higher value product for various purposes as mentioned in the previous chapter of this study.

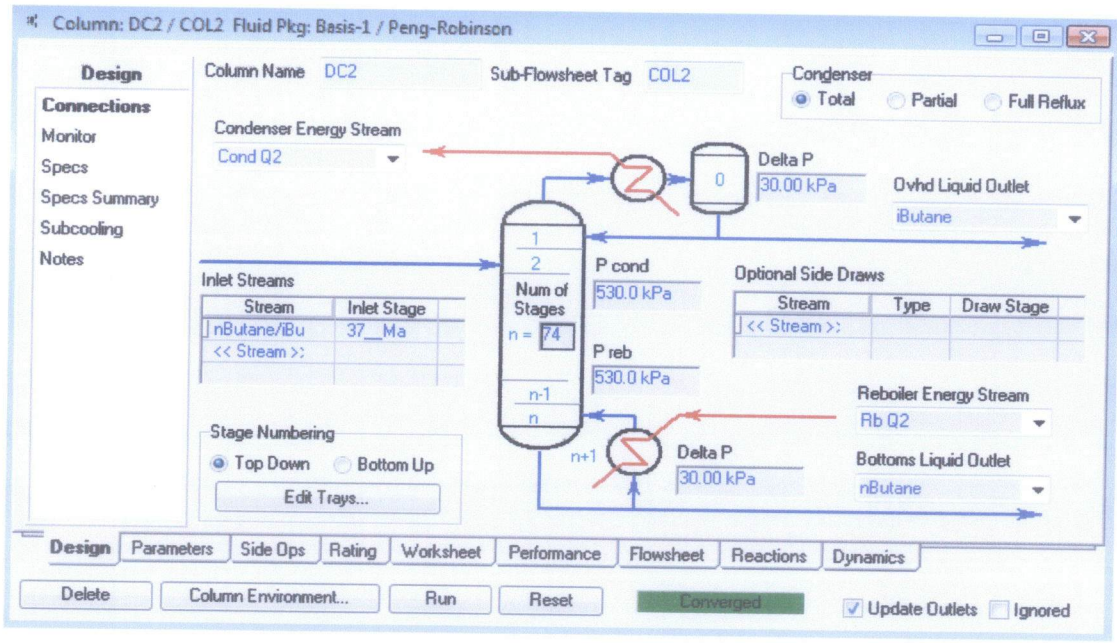

Figure 26: Base case simulation for DC2

Table 7: Product specifications

Components	Mole Fraction	References
Isobutane	0.970	(Puyang Lianzhong Xingye Chemical Company Limited, 2011)
Propane	0.900	(Propane 101, 2011)
Normal butane	0.975	(Puyang Lianzhong Xingye Chemical Company Limited, 2011)

The overview of the base case simulation is illustrated in Figure 27.

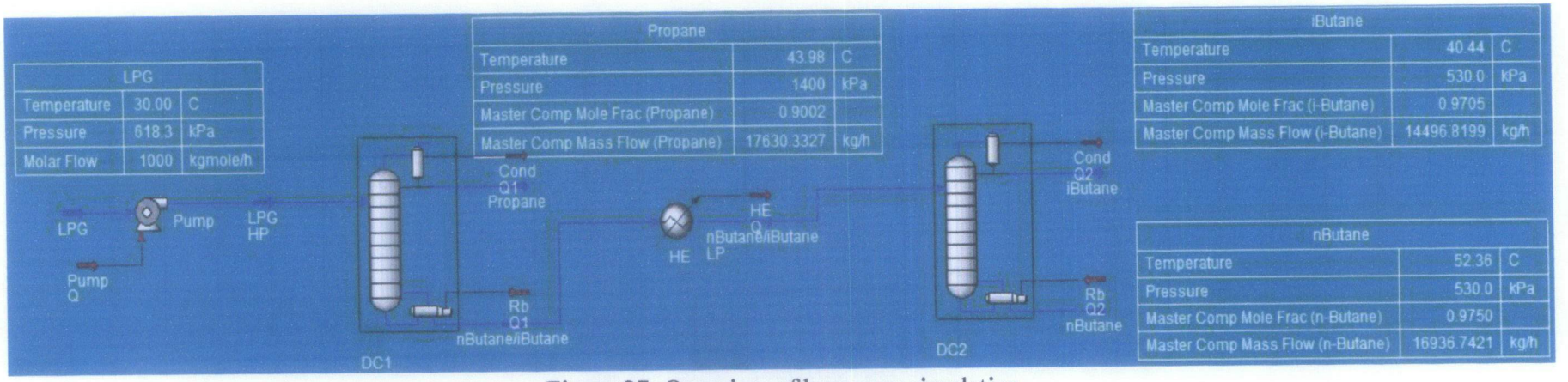

Figure 27: Overview of base case simulation

3.4 Simulation of Indirect Sequence

The feed enters the first distillation column with a pressure of 1000 kPa and temperature at 30.36 °C. The feed was entered at 39th stage of the distillation column. In the DC1, the pressure was simulated at 1000 kPa which is lower than the base case. In the indirect sequence process, n-butane was separated first at the bottom of DC1. The top product which consists of isobutane and propane exits the DC1 at 40 °C. At the bottom of DC1, n-butane exits at 79.08 °C, has the desired purity of 97.0% of mole fraction, similar to the base case, which is suitable to be used as petrochemical feedstock. In DC1 for this case, the number of stages required was 75 trays for the simulation to converge. Higher number of trays was required as compared to the base case. This is due to the smaller difference in relative volatility between the two products. The reflux ratio for this case to converge was 4.8. Figure 28 shows the configuration of DC1 for the indirect sequence.

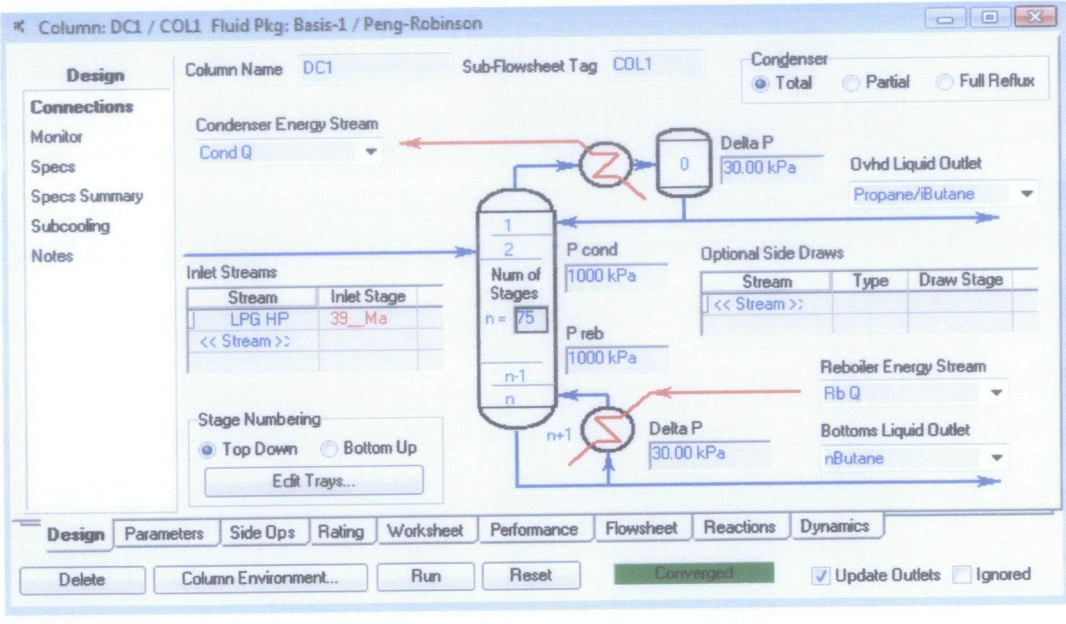

Figure 28: Indirect sequence for DC1

The streams that contains isobutane and propane from the top product of DC1, has to be increased to 1400 kPa as compared to the base case that was configured to be at 1000 kPa, before entering DC2 as it affects the vapour pressure of the contents in the column and thus, enabling it to converge. The feed entering the column contains propane and i-butane at 58% and 42% of mol fraction respectively. The amount of n-butane is very small as compared relatively to propane and i-butane which is about 0.1%. This is an indication that most of the n-butane has been separated in the DC1. The reflux ratio simulated was 1.22 and with 34 trays. The feed from the DC1 was entered in the 18th trays. The quality of the product was able to be kept constant for the process simulation of indirect sequencing. Sequencing the process indirectly, does not remove the light key component first, is expected to demand higher duty and energy load for the same product specification as compared relatively to the base case. This will further be illustrated and discussed in the next chapter of this study. The configuration of the DC2 for the indirect sequence is illustrated in Figure 29. The overview of the whole case for the indirect sequencing is illustrated in Figure 30.

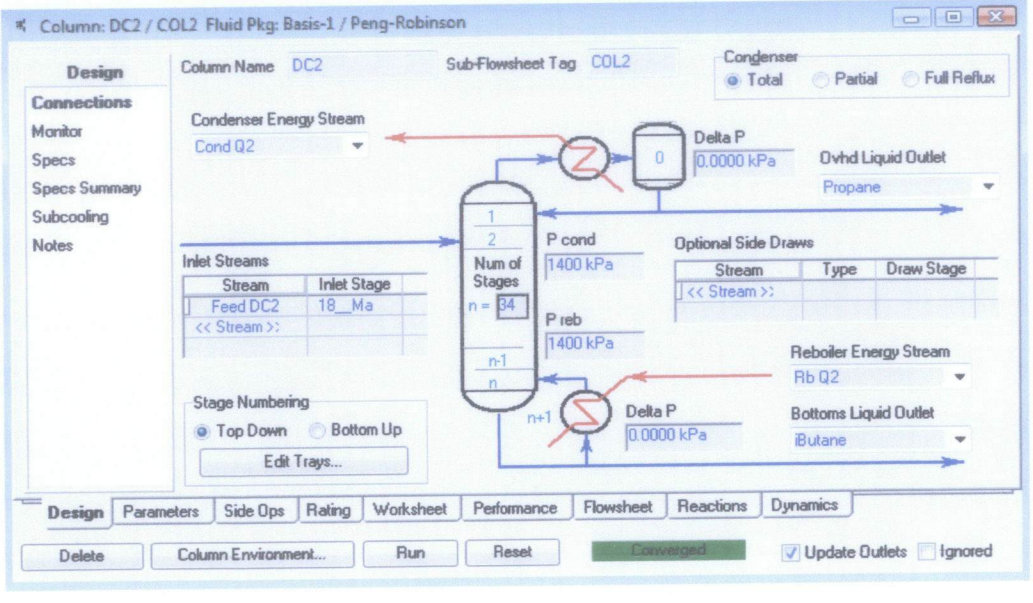

Figure 29: Indirect sequence for DC2

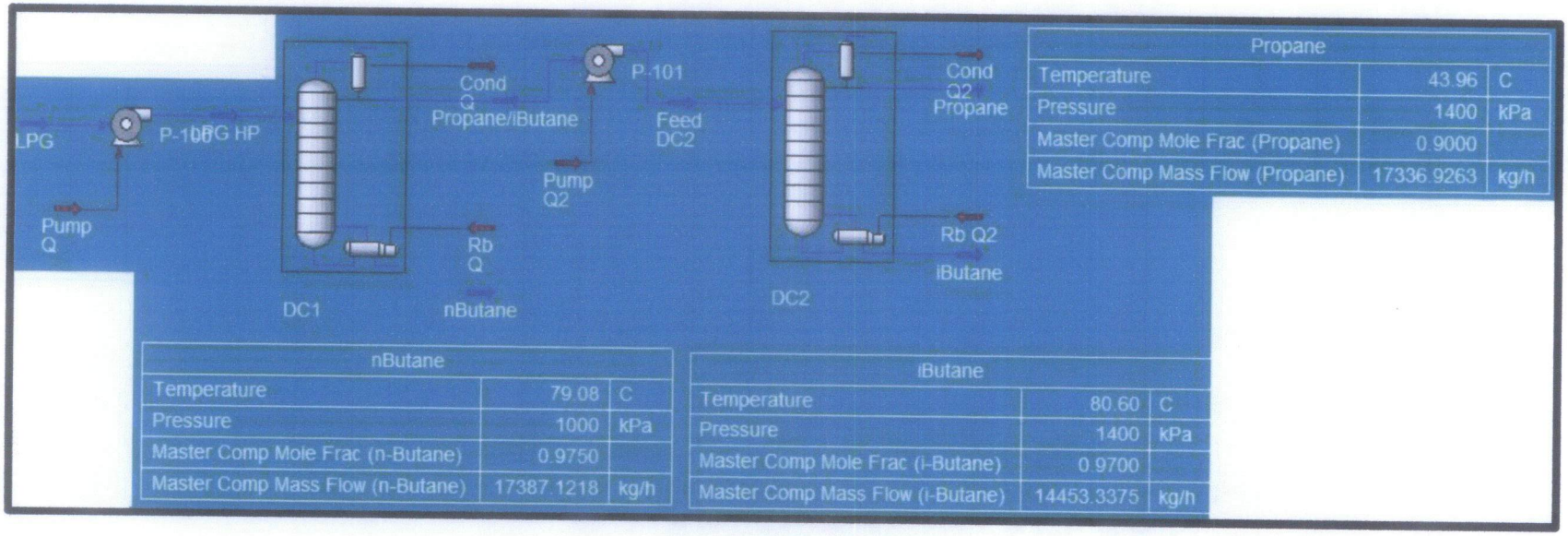

Figure 30: Overview of indirect sequence simulation

3.5 Simulation of Dividing Wall Column (DWC)

Similar to the direct sequence (base case) sequencing, the LPG feed is fed into a pump to increase the pressure to 1400 kPa. Similar reason with the base case, this will ensure that cooling water can be used condense the top product of the dividing wall column. The feed that exits the pump then enters a vessel which acts as a prefractionator section of the column. Figure 31 illustrates the setup of the vessel which acts as the prefractionator part of the DWC. In Aspen HYSYS simulation software, there is no equipment or tools for DWC. Thus, the application of DWC can be conducted through the setup of separate columns however; in this case, a vessel was used as the prefractionator section of the DWC. Results obtained from the simulation using Aspen HYSYS for DWC configuration are reliable (Ignat & Woinaroschy, 2011). In the prefractionator, the duty supplied to create pressure difference of 50 kPa between the top product and the bottom product. About 9.921 kW of duty were required.

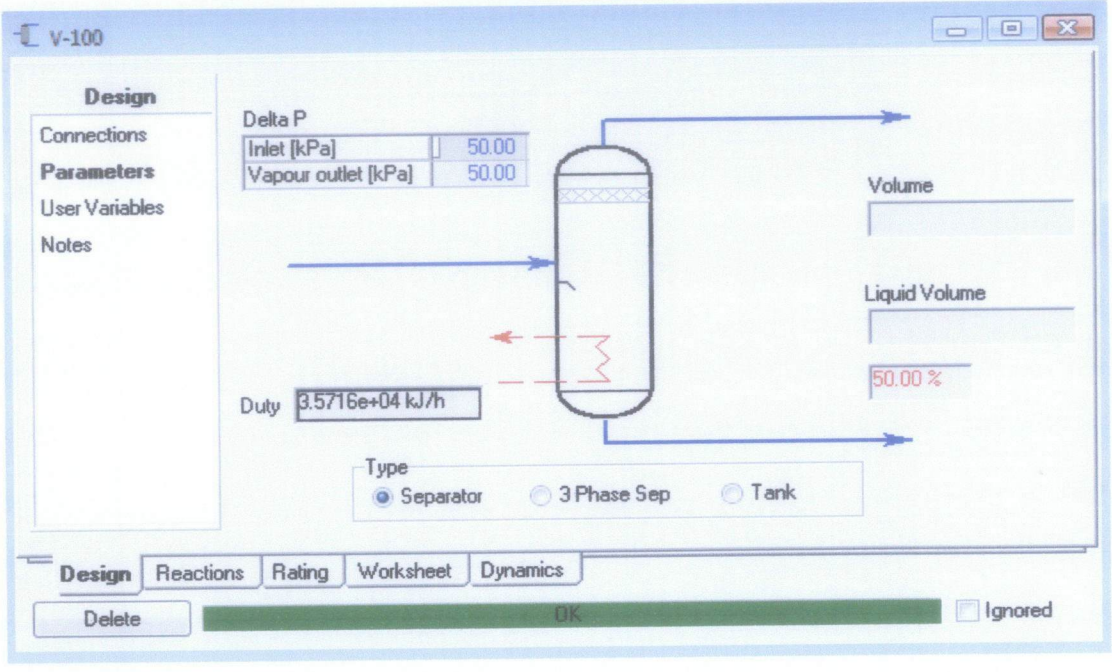

Figure 31: Prefractionator section of the DWC

The prefractionator has enabled initial purification of the constituents of LPG to be attained. Figure 32 and Figure 33 shows the composition of the top and bottom

stream exiting the prefractionator section of the column (in this case, the vessel). The top product has shown changes as compared to the LPG composition. This indicated that the pressure difference created in the vessel has performed some degree of preliminary purification before entering the column. However, the bottom product appears to have similar composition as the LPG supplied. This is likely due to the limitation that the amount of purification that the prefractionator is able to perform. Large portion of the LPG will be purified by the column as the main purpose of a prefractionator is to perform preliminary purification, reducing remixing and re-separation.

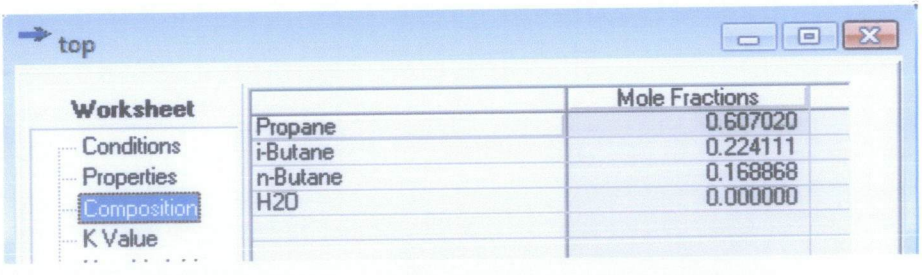

Figure 32: Top product of the prefractionator

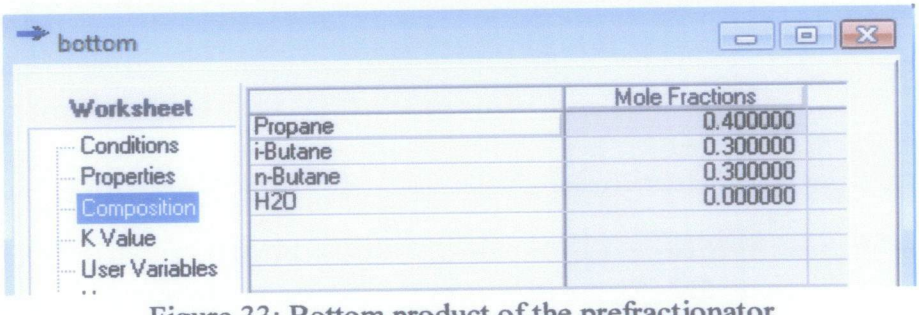

Figure 33: Bottom product of the prefractionator

In the column of the DWC, a distillation package was used consisting of a reboiler and a condenser. Figure 34 shows the setup of the column section of the DWC. The number of stages for this configuration was varied between 60 stages to 100 stages. Consistent with the theory, the bigger the number of trays the lesser the duty required for the condenser and the reboiler. This is because the reflux ratio decreases with the increase in the number of stages of the column. All of the 3

components were able to achieve the specifications required. Since propane is the lightest key, it will be the top product from the column. i-butane will be extracted from the 38th stream and n-butane will be the bottom product.

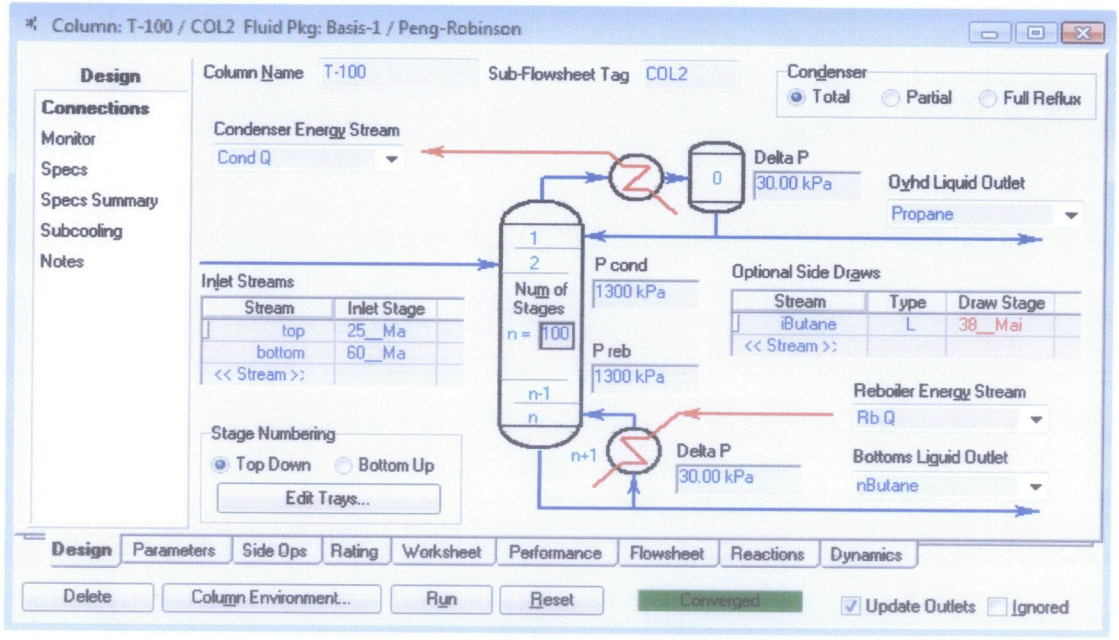

Figure 34: Column section of DWC

The overview of the DWC configuration is shown in Figure 35.

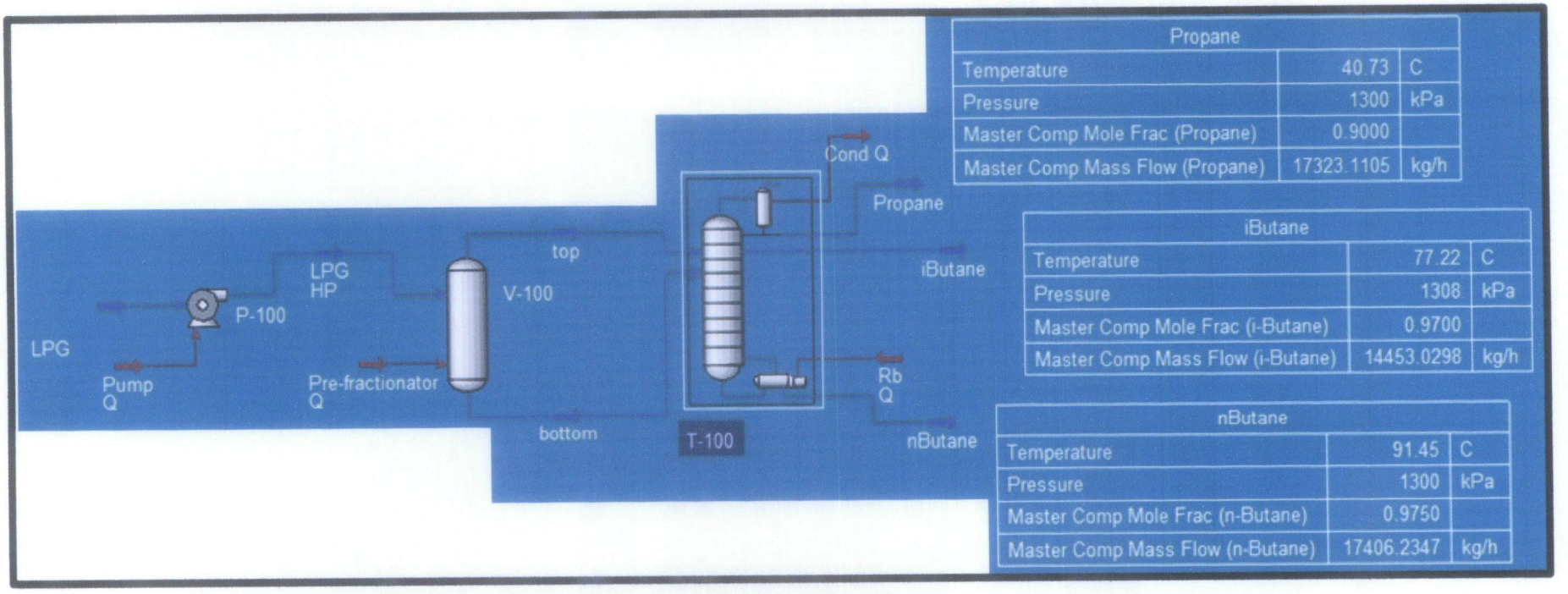

Figure 35: Overview of the DWC column

CHAPTER 4

RESULTS AND DISCUSSION

In determining the feasibility of the conceptual design, it is important that the separation processes of the conceptual design simulated varies so that a more vigorous analysis and alternatives are studied. This study has successfully converged 3 configurations namely, direct sequence (base case), indirect sequence and dividing wall column. The 3 configurations were able to purify the LPG constituents to the required purity (refer to Table 7) that is suitable to be utilized for petrochemical feedstock. It is crucial that the development of alternative methods in acquiring the similar aim is studied becausean optimal design which uses the least energy for the separation of LPG constituent can be obtained. Also, as more alternative configurations are successfully developed, it will increase the basis of feasibility of this study.

Table 8illustrates the total amount load and the duty required for the separation processes of LPG. The base case consists of condenser duty, reboiler duty, pump load and cooling duty. The feed to DC2 in indirect sequence, pump is required to increase the pressure to 1400kPa so that the conditions for convergence can be met. Between the three alternatives developed, the direct sequence utilizes the least amount of energy in terms of condenser duty and reboiler duty. The pump load differences are very small and they are due to the lower inlet feed pressure required prior entering into the DC1 for the indirect sequence.

It is expected that the indirect sequence would require higher amount of energy as compared to the direct sequence. This is because propane which accounts about 40% of the total composition of the LPG, is the largest constituents as compared to i-butane and n-butane which is about 30% each. Therefore, in indirect

sequencing as compared to direct sequencing, removal of n-butane first in the bottom of DC1, resulted in the top stream having larger amount of load to be fed into DC2. Regardless propane being the lightest key and having large relative volatility as compared to i-butane and n-butane, it most likely have the largest load effect as it comprise the largest fraction in the LPG. However, the difference in terms of condenser and reboiler duty are just about 2 093 kW and 3 715 kW respectively.

The dividing wall column utilizes the most amounts of energy condenser duty and reboiler duty, contrary to what have been discussed in the literature review chapter. Relative comparison with the direct sequence case, the amount of energy for condenser duty and reboiler duty are about 4 times more than the direct sequence. However, the prefractionator section of this column, only 9.92 kW of energy was required for the initial separation. However, the energy utilization for DWC could be improved through optimization in the DWC configuration such as the changing the type DWC configuration, exit streams and recycle stream that are not in this scope of this study.

Table 8: Duty and load requirement summary

	Direct Sequence (Base Case)	Indirect Sequence	Dividing Wall Column (DWC)
Condenser duty (kW)	17 953	21 992	77 150
Reboiler duty (kW)	19 634	23 349	78 670
Pump load (kW)	28.58	24.26	28.58
Cooling duty (kW)	1 084	-	-
Pre-Fractionator (kW)	-	-	9.92

Estimation in terms of product value was also calculated and shown in Table 9. It can be observed that all the 3 components produced by the successful simulations generate a product value which is close. The total product value of all ranges from RM 41,000- RM 42,000. Thus, the estimated value would be insufficient

to determine which process simulation would be better as the difference is not significant.

In order to evaluate the separation processes efficiency, the throughput of the converged cases simulated can be done. Of the three components in LPG, isobutane which has the highest value when compared relatively to n-butane and propane. However, all of the constituents of LPG must be taken into consideration as thorough evaluation and optimal value creation of all products will lead to a more feasible study. Table 9 shows the estimated for each case of separation processes for each constituent. The detailed calculation is attached in the Appendix 1. It can be observed that the direct sequencing which is the base case in this study gives slightly higher value than the indirect sequencing and the DWC. However, when only duty and energy are taken into considerations, it can be observed that the base case is the more economical choice as base case will be lower in terms of operating costs due to lower condenser duty and reboiler duty in spite of the need for cooling duty. Nevertheless, it is important that costs of capital in setting up the facilities be evaluated as well.

The split of propane first that has resulted in the removal of small fraction of i-butane and n-butane as the purity of propane required is just 90.0% of mole fraction. The remaining 10.0% will constitute of i-butane and n-butane. Table 10 shows the summary of throughput base on the price of purifying the LPG constituents. The direct sequence has the highest throughput. Thus, it can be claimed that for this study, the direct sequence would be the best option in terms of throughput as it 1.267 kg of product purified per kW of energy used.

Table 9: Estimation of product value for each case

Product	Density (g/cm^3)	Price (RM/gal)	Direct sequence separation amount (kg/h)	Indirect sequence separation amount (kg/h)	Dividing wall column separation amount (kg/h)	Base case product value (RM/h)	Indirect sequence product value (RM/h)	DWC sequence product value (RM/h)
Isobutane	2.51	4.64	14497	14453	14450	7071.04	7049.58	7056.60
N-butane	2.48	4.46	16937	17390	17397	8036.75	8251.70	8264.97
Propane	0.493	2.86	17630	17335	17330	26985.61	26534.06	26558.36
		TOTAL	**49064**	**49178**	**49177**	**42093.40**	**41835.34**	**41879.92**

Table 10: Throughput summary

	Direct sequence (kg/kW)	Indirect sequence (kg/kW)	DWC (kg/kW)
Throughput	1.267	1.085	0.315

As for the cooling medium, both the base case and the direct sequencing are in the range where cooling water can be used to cool the medium. Likewise, reboiling and heating up of the streams which is less than 100°C, allows the usage of steam. Therefore, the 2 cases does not require cooling or heating with the use of additional utility or expensive medium such as oil or chilled water. The cooling medium is important aspect to be evaluated as it will affect the operating costs of a plant. This conceptual design will be more feasible as the operating costs can be minimized.

The capital expenditure of the processes will depend much on the design of the conceptual plant. Therefore, Table 10 illustrates relative comparison on which processes would likely require the least capital expenditure. The higher the number of trays will lead to a higher capital expenditure as the column will have to be built tall and inserted with large number of trays. As for the pressure, higher pressure would result in higher costs of construction for the equipment, thus leading to higher capital expenditure. The number of trays for both of the system is almost similar except that the number of trays for indirect case is higher than the base case for the DC1 as it is separating the n-butane first. Due to that, the number of trays in the DC2 for the indirect sequence is lower when compared to the base case.

As for the DWC, the number of stages was varied from sixty to one hundred. Consistent with the literature reviewed, as the number of stages is increased, the reflux ratio decreases. A decrease in reflux ratio will lower the energy requirement for the condenser and the reboiler. In this simulation for the case of DWC, the number of stages was chosen to be 100, in spite of it being a very large number of stages, is because at low number of stages, for example 60 stages, the amount of

energy required for the condenser and reboiler were too large. When comparison were made with the other cases, the condenser duty, for example is 10 times more than the base case. However, in order to improve the results of this analysis, it is important that the DWC configuration is studied in depth which will be discussed in the later part of this study, for future works and recommendations.

From Table 11, the capital expenditure (CAPEX) of DWC would most likely be the lowest as compared to direct sequence and indirect sequence. However, by reducing the number of stages, as mentioned previously, it will further reduce the CAPEX required to set up the plant but this will then increase the operating expenditure (OPEX) of the DWC. The increase in duty for the reboiler and condenser will demand more energy which will directly affect the operating costs. Therefore, it is crucial that the optimal trade-off between the CAPEX and OPEX be found. Aside from that, the DWC which only requires one column is another advantage in terms of CAPEX when compared to the direct and indirect sequencing. The need to only design and construct one column will definitely incur lesser costs. However, the design of the DWC is much more complicated as compared to the conventional columns.

As for the direct sequence, it will be more in favour when compared to the indirect sequence. With almost similar total number of stages, the pressure of the column will likely be the factor which will affect the CAPEX for this case. Higher pressure operations will require more expensive materials for construction as it has to withstand the higher pressure. Thus, in terms of CAPEX, based on the pressure of DC2 of direct sequence which is just 530 kPa, it will require cheaper material for construction, leading to a lower CAPEX.

Table 11: Summary on the number of trays and pressure specification

Case	DC1 Pressure (kPa)	DC1 Trays	DC2 Pressure (kPa)	DC2 Trays	Pre-fractionator
Direct sequence (Base case)	1400	34	530	74	-
Indirect sequence	1000	75	1400	34	-
Dividing wall column	1300	100	-	-	Required

CHAPTER 5

CONCLUSION AND RECOMMENDATIONS

5.1 Conclusion

The result obtained for direct sequence (base case), indirect sequence and DWC provides a benchmark for which could be improved when future work are executed. The purity of the products separated should have an optimal throughput that is producing the most products while lowering the amount of energy used. For the successful results obtained from the converged simulations, the energy load and duty have been identified. The ideal situation is to produce high amount of throughput while maintaining the purity of the product so that it is adequately pure to be used as feedstock. The products that are not used as chemical feed stock can be mixed back and be used as LPG and consequently can be sold for various purposes for example for domestic use. This will ensure that there is no waste and profit can be optimized. This study has provided an insight of the preliminary feasibility of creating additional value from LPG. With the base case, indirect sequence and DWC producing about RM 41 000 to RM 42 000 per hour can be used as a basis on the potential return on investment (ROI) if this study if pursued further in depth. As for the throughputs, results have indicated that, at the moment, the direct sequence is preferred as its throughput is the highest with 1.267 kg of products purified per kW of energy used as compared to the indirect and DWC case with throughputs of 1.085 and 0.316 of purified constituents per kW of energy consumed.

5.2 Recommendations and Future Works

An aspect that can be studied for further analysis is to evaluate on acquiring the optimal operating expenditure and capital expenditure. An in depth feasibility study can be conducted to perform the CAPEX and OPEX for a certain duration so that realistic economic feasibility can be calculated. By doing so, the feasibility of establishing a plant for this project can be evaluated more thoroughly.

Aside from that, it is hopeful that further optimization of alternate processes aside from the 3 cases done, can be simulated as well. By doing so, this study will be able to provide more alternatives and more feasible conceptual separation processes that would be important as an indication of potential business prospects in the separation or purification of constituents in LPG in Malaysia to create added value. An area for further studies and possibly, performing similar simulation to acquire the desired specifications of products can be done would be to perform other configuration of DWC. In this study, the Petyluk configuration has been executed. However, in DWC, there are numerous configurations that can be conducted and studied that would enable DWC to produce a higher throughput as suggested by many literatures.

An in depth evaluation of the project feasibility could not be conducted in this study because of the scope of this study which does not cover areas such as realistic OPEX and CAPEX evaluation. In spite of that, this study has enhanced the potential to create value and is hoped to provide the motivation required for potential researchers to scrutinize on this subject further. It is crucial that the detailed economic study which includes capital expenditure and operating expenditure should be conducted. If pursued further on this topic, the amount of expenses required to set up the plant shall be weigh against the value gained from the purified products. The result of the mentioned comparison will enable a firm indication on whether the separation of LPG will actually be a feasible and implementable conceptual design.

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